

“Waste-Heat Powered Ammonia Absorption Refrigeration Unit for LPG Recovery”
FINAL REPORT
DOE Contract DE-FG36-03GO13128

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Formerly Giant Industries Inc.
Bloomfield, New Mexico Refinery
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EXECUTIVE SUMMARY

An emerging DOE-sponsored technology has been deployed. The technology recovers light ends from a catalytic reformer plant using waste heat powered ammonia absorption refrigeration. It is deployed at the 27,000 bpd Bloomfield, New Mexico refinery of Western Refining Company.

The technology recovers approximately 50,000 barrels per year of liquefied petroleum gas that was formerly being flared. The elimination of the flare also reduces CO₂ emissions by 17,000 tons per year, plus tons per year reductions in NO_x, CO, and VOCs.

The waste heat is supplied directly to the absorption unit from the Unifiner effluent. The added cooling of that stream relieves a bottleneck formerly present due to restricted availability of cooling water. The 350F Unifiner effluent is cooled to 260F. The catalytic reformer vent gas is directly chilled to minus 25F, and the FCC column overhead reflux is chilled by 25F glycol.

Notwithstanding a substantial cost overrun and schedule slippage, this project can now be considered a success: it is both profitable and highly beneficial to the environment. The capabilities of directly-integrated waste-heat powered ammonia absorption refrigeration and their benefits to the refining industry have been demonstrated.

GOALS AND PROJECT OVERVIEW

Goals

One goal of this project was to deploy an emerging technology, namely to conduct a second field demonstration of a DOE-sponsored patented light ends recovery technology at an oil refinery. From the perspective of the host refinery, the goal is to recover light ends (LPG) sufficient to extinguish the flare, using waste heat to power the recovery system.

Technology

Many refinery processes produce “light ends” (propane and lighter molecules). Much of this gas is used for hydrotreating or as fuel gas, but the amount produced usually exceeds these needs. Compared to the alternative of flaring the excess light ends, it is very desirable both from efficiency and economic perspectives to recover the excess. That is done by “light ends recovery” technology, also referred to as LPG recovery. There are a variety of methods by which light ends can be recovered. They include: lean oil absorption; mechanical refrigeration; molecular sieve adsorption; and membranes. All of these traditional methods are both capital-intensive and energy-intensive. Almost all large oil refineries use lean oil absorption on the particular stream this project has targeted. However the lean oil absorption technology is too costly at small scale, so most small refineries do not recover those particular light ends. This results in many of them having routine flaring.

The light ends recovery technology being demonstrated uses refinery waste heat to produce refrigeration using an ammonia absorption refrigeration cycle. By directly integrating the absorption refrigeration unit with the catalytic reformer plant, the capital cost is kept low. Use of waste heat to produce the refrigeration keeps the operating cost low.

The ammonia refrigeration directly chills the net gas and/or treat gas from the catalytic reformer plant sufficiently to condense out the bulk of the C3 (propane) and heavier fractions from that gas. Whereas those fractions only constitute about 8% of that gas by volume, they are more than 40% by weight, and also by heat content. When those light ends are not removed, the refinery is typically in an excess fuel gas condition, wherein the excess fuel gas is flared. When they are recovered, the flare is reduced or eliminated. Thus the recovered light ends are obtained without any increase in feedstock to the refinery – they are in effect recovered from the flare. When the light ends recovery is pushed to the point that the refinery is in a fuel gas deficit situation, it is necessary to import natural gas to make up the deficit. Even that is a beneficial outcome, since the market value of the light ends is appreciably higher than natural gas, and the conversion of natural gas to light ends is at essentially 100% efficiency.

The amount of waste heat available in the catalytic reformer plant greatly exceeds the amount necessary for light ends recovery. Hence once the refrigeration plant is justified by the economics of light ends recovery, the marginal cost of adding capacity for other refrigeration needs is low. Whereas refineries have traditionally made very little use of refrigeration (because of the high cost of producing it), it turns out that when it is nearly free, numerous beneficial uses can be found. This project included one such use: chilling the overhead reflux to the FCC separator column, to increase its recovery.

Previous Demonstration

This waste heat powered light ends recovery technology was first demonstrated at the Suncor Commerce City refinery in Colorado. The plant was commissioned in 1997, when the refinery was owned by Total, and operated for eight years, during which time the refinery ownership changed to Diamond Shamrock, then to Valero, and finally to Suncor. When the refinery recently transitioned to a highly-deficit fuel gas condition, the recovery unit was shut down. More recently, a major turnaround has placed the refinery back in an excess fuel gas condition, and hence it is planned to re-commission the absorption refrigeration unit.

This first demonstration of the light ends recovery technology provided a proof of the concept and of the technical feasibility. The original capital cost estimate showed a two-year payback on the investment when oil cost was less than \$20 per barrel. The installed cost ended up at something near twice the initial estimate. However, by that time oil price had increased to over \$40 per barrel, so the two year payback was still realized. The US Department of Energy provided a 30% cost share for that demonstration plant.

Several lessons were learned pursuant to the previous demonstration. With regard to major components, initially two of them initially under-performed – the absorber and the desorber. Both of those units have very unusual and exacting requirements: temperature glide; concentration glide in both liquid and vapor phases; mass exchange between phases; and

countercurrent heat exchange. Having mass exchange between phases on both sides of the transfer surface further complicates the desorber. The absorber is further complicated by having to operate at below atmospheric pressure, such that low pressure drop is paramount.

The vendor used commercial software to design the shell and tube desorber. This demonstration showed that the software is not up to this demanding task.

The absorber was designed by Energy Concepts. The first design, a welded plate configuration, had channeling problems, and had to be replaced. The second design, consisting of coiled tubing and a perforated plate distributor, provided close to design performance. Fortunately, the second refrigeration load designed into this plant (chilling the wet-gas compressor inlet vapor) turned out to be unnecessary. Hence that excess capacity was used to supplement the light ends recovery refrigeration, thus overcoming the deficiencies of the desorber and absorber and achieving design recovery.

Objectives of the Emerging Technology Deployment

The objective of this project was to deploy an emerging technology that had been developed with cost-share assistance from the U.S. Department of Energy. This deployment demonstrated the technology in a different refinery setting, recovering nearly 50,000 barrels per year of LPG while largely eliminating the flare. The technology converts low level waste heat into chilling, which supports the Petroleum Vision Roadmap goals of “innovative technology for recovery of low level waste heat”. Hence this project further publicizes this promising energy efficiency technology to the refining industry. The U.S. Department of Energy provided a 30% cost share. As discussed later, cost escalation caused that share to decrease to about 14%. The design improvements that were verified in this second demonstration project included a new cycle, new components, operation with two separate refrigeration loads, and improved operability.

Actual Accomplishments Compared to Objectives

The new LP absorber (a shell and tube type) provides acceptable performance – the pressure drop is very low (less than 1 psid), and the overall U value is about 80% of design.

The IP absorber (a shell and coil type) provides excellent performance – the pressure drop is somewhat higher (approximately 2 psid) but the overall U value is about four times higher than that of the LP absorber. Assuming that its performance doesn't degrade too much as it transitions to operation in vacuum, that design should be tried out as the LP absorber as well.

Several other new components, including the rectifier and the refrigerant heat exchanger, also showed acceptable performance.

Although the system is achieving its design low refrigeration temperatures, it is currently under cold weather conditions, and at reduced loading. The system has not yet been tuned for optimum operation, and that will likely be required in order for it to achieve design performance at peak summer conditions.

From the controls and operability perspective, some improvement has been demonstrated relative to the first plant, but further improvement is definitely warranted. This plant operates smoothly once it is settled out, but startup remains lengthy and more laborious than it should be. The technology developer can implement and test revised startup procedures that would resolve the startup difficulties. However there has not been such an opportunity on this plant as of yet.

The system was provided to the site as three skidded units. Photo 1 shows the absorption skid ready for shipment. Photo 2 shows the installed configuration of the hydrocarbon skid (on the right) and the absorption skid. Photo 3 is another view of the installation taken at night, showing the evaporative cooler skid.

PROJECT ACTIVITY SUMMARY

Project Chronology

This DOE sponsored cost-shared project commenced with a proposal to a DOE RFP in November 2002, and a contract award in September 2003. The original plan was to conduct the deployment at the Giant refinery in Yorktown, Virginia. However, a few months into the detailed design the Yorktown refinery decided to install a new hydrocracker unit. That unit would consume all available light ends (which previously were flared). Also, at that location natural gas cost more than LPG. Hence the rationale for the light ends recovery unit disappeared.

Within a few months, it was determined that the light ends recovery would benefit the Giant refinery in Bloomfield, New Mexico because it also had a continuous flare. Hence in September 2004 the DOE contract was modified to reflect the change in demonstration site, and a new detailed design was initiated.

A few months later, in early 2005, the Bloomfield refinery started experiencing severe feedstock shortages, and was very uncertain about future crude availability. Accordingly work was halted once again. It wasn't until December 2005 that the crude shortage was resolved and the project received a formal go-ahead.

Detailed design and skid fabrication proceeded during the first seven months of 2006, plus some preliminary site preparation work. The installation of the three skids comprising the overall system was scheduled for one month starting in August, but ended up taking two and a half months. Finally, in November 2006, the system was ready for instrument checks, charging, startup, and commissioning. That process, which should have taken at most about three weeks, ended up taking about four months, for a variety of reasons, some of which are enumerated below.

The system operated at highly non-optimal conditions throughout the summer of 2007, providing only about half of design LPG recovery. The technology developer was brought in for one week in November 2007, and made a couple of critical adjustments that brought the unit up to design performance.

Technical Approach

This project uses an ammonia absorption refrigeration cycle, which is capable of both the required low temperatures (-30°F) and of waste-heat powering. The input heat is 350°F liquid effluent from the Unifiner, which otherwise must be cooled with cooling water. The low temperature (-30°F) refrigeration is applied to the catalytic reformer plant net gas (also called vent gas). The intermediate temperature refrigeration (+25°F) is applied to FCC plant overhead reflux, via a closed loop glycol system. Figure 1 is a simplified process flow diagram of the system.

Both the driving heat and the refrigeration heat must be rejected from the absorption cycle. Part of that heat is rejected in a evaporative condenser, where 150 psig ammonia vapor condenses to liquid. The remainder is rejected at the two absorbers (LP and IP), via an aqua cooling loop. The aqua coolant in turn is cooled in an evaporative cooler. Both the evaporative cooler and the condenser are incorporated in a single wet surface air cooler, which has VFD fan speed control to maintain constant condensing pressure.

Project Results

A computerized data acquisition and control system continuously records the readings from 19 instruments. Beyond that, there are approximately 60 additional local instruments with manual readouts. The automatic readout instruments are sufficient to assess the overall results of system operation (e.g. LPG recovery rate). A full analysis of system operation down to component level requires that all of the readings be recorded at steady state conditions. This is a very demanding requirement. It takes the system approximately one hour to settle into full steady state (no more small changes in levels or concentrations). Then it takes approximately an hour to record all the manual readings. Thus if either refinery operations change or ambient temperature changes during that two hour window, the steady state is invalidated, and must be started again. One result of this is that only three steady states have been recorded thus far.

Table 1 is a printout of two hours of data (5-minute intervals) from the “Procidia” data recording system, on November 8. Note there are 6 flow readings, 4 pressures, 8 temperatures, and one level (LPG separator).

Figure 2 is the graph of one of those readings (LPG flow) for the entire month of November. Figure 2 shows that the LPG recovery started the month at about 80 barrels per day, which is the value it had been at for most of the summer. The technology vendor arrived November 5 and made some adjustments, which within 24 hours increased LPG recovery to the design value of 150 barrels per day. Shortly thereafter, the refinery transitioned into winter-operating mode, with substantially reduced feed rate. Hence the recovery declined to about 100 barrels per day, remaining there through the winter. The LPG temperature remained at about -20°F, showing that the system was meeting design.

Table 2 presents the three recorded steady states – both Procidia readings and manually recorded readings. The technology provider recorded the first two steady states during the maintenance visit from November and the third steady state was recorded by refinery personnel.

Table 3 presents the results of the system analysis for the above three steady state data sets. The heat duty and performance of each major heat exchanger is determined. Overall cycle performance is also calculated (both tonnage and COP). The COP (Coefficient of Performance) of 0.37 is typical of double lift cycles, and right at the design value.

Noteworthy items:

1. The 11/15 run did not have any intermediate temperature refrigeration load, due to conditions in the FCC plant.
2. The IP letdown outlet quality should be zero. The 10% or so quality was due to a malfunctioning shutdown solenoid valve (not receiving open signal), with manual bypass valve set for zero level.
3. The LP pressure of 0.1 psig corresponds to an absolute pressure of 12 psia at the Bloomfield altitude (5,200 ft). With an assumed 1.5 psid drop between the LP evaporator and the LP receiver, this corresponds to an LP consistent temperature of -32°F. That is consistent with the recorded LPG temperature of -22°F.

Economic Performance

Given the schedule slippage and cost escalation problems described below, the installed cost escalated from the original estimate of about \$1.9 million to a final actual cost of \$3.4 million. As a result, the fixed DOE cost share decreased from 30% to 14% of the project total. On the benefits side, a rough order-of-magnitude calculation of the value of the LPG recovery is as follows:

- 150 barrels per day for 8 months
- 100 barrels per day for 3 months
(one month of turnaround and other maintenance)
- Annual LPG recovery: 49,500 barrels
- Value, at an assumed price of \$80 per barrel: \$3,960,000

This shows that there is strong financial incentive to keep this recovery unit operating at full capacity.

The question arises as to how much natural gas cost should offset the above savings, for those times when flare reduction alone doesn't yield the full 150 bpd. That is difficult to calculate, but it raises one important issue: when the refinery is in a fuel gas excess situation, and hence flaring, there is little or no incentive to repair heat leaks, steam leaks, etc. However, once it goes into a fuel gas deficit situation, where natural gas is being imported, there suddenly is a financial incentive to fix those leaks, and many are both easy and inexpensive to fix.

Problems Encountered and Lessons Learned

1. Operation

The brief full-capacity operation in the early part of November demonstrated that both design refrigeration temperature (-25F) and design LPG recovery (150 barrels per day) are achievable with this plant.

That good level of performance still needs to be demonstrated at design summer conditions (warm ambient). Two points are noteworthy in that regard:

- a. The catalytic reformer plant characteristically has more light ends in the net gas at warm ambients, compared to cold. Hence it is likely that the LPG recovery will increase in the summer, even if the -25°F temperature isn't fully achieved then.
- b. In order to achieve optimal performance of the ARU (and hence that low refrigeration temperature), five flow splits must be optimized by adjusting manual throttle valves. That is an iterative process requiring several hours at steady state conditions plus intimate knowledge of the system. It is recommended that the technology developer do this in the late spring of this year.

2. Management

Early in 2006 it was decided to bring onboard an Engineering, Procurement and Construction (EPC) firm experienced in refinery projects. They were to be responsible for the design of the hydrocarbon skid, to ensure overall code compliance, and to supervise the equipment installation at the site. The rationale was that their related experience would help ensure against schedule delays and cost overruns. Unfortunately, those objectives were not achieved. In retrospect, it is believed that two factors contributed to that. First, this EPC did not have any significant knowledge of ammonia absorption technology (very few if any do). This directly led to some avoidable problems. Secondly, and more importantly, the management structure evolved to the point where the technology provider was effectively removed from participation in management decisions, particularly after installation began. Once again, this led to easily avoidable problems. This management structure could have and should have worked, but only if there was close cooperation between the EPC, the technology developer, and the refinery management.

3. Pumps and Accessibility

The absorption skid requires three pumps plus 100% installed spares (six pumps total). The pumps were originally on a subassembly confined within the skid. During installation, the pump subassembly was relocated outside the skid confines, to improve access. This added cost and time to the installation process. Future skid designs will incorporate convenient pump access.

4. Relief Valve Issues

The ammonia solution containment is divided into four sections, each with its own safety relief valve. Late in the design process it was determined that the governing relief event in three of those sections was rupture of a tube in the associated hydrocarbon heat exchanger. Hence those relief valves had to be upgraded to the size necessary for hydrocarbon relief duty, including all associated features (rupture diaphragms, discharge pressure sensors, etc.). Also, the relief valve discharges had to be piped to the hydrocarbon flare. This was further complicated by the need to avoid the possibility of standing liquid heads in the relief discharge piping.

5. Other Miscellaneous Issues

The computerized control system was sending phantom and spurious shutdown signals to the system for the first several months of operation. Closer communication and cooperation between the control system developer and the technology developer can avoid recurrence of this problem.

Several pipe fitup problems on the absorption skid had to be repaired, requiring an extra week of installation time. Shop procedures that allowed those problems have been corrected.

The drain fittings and thermocouple wells on the hydrocarbon heat exchangers were originally threaded, and had to be changed to all-welded.

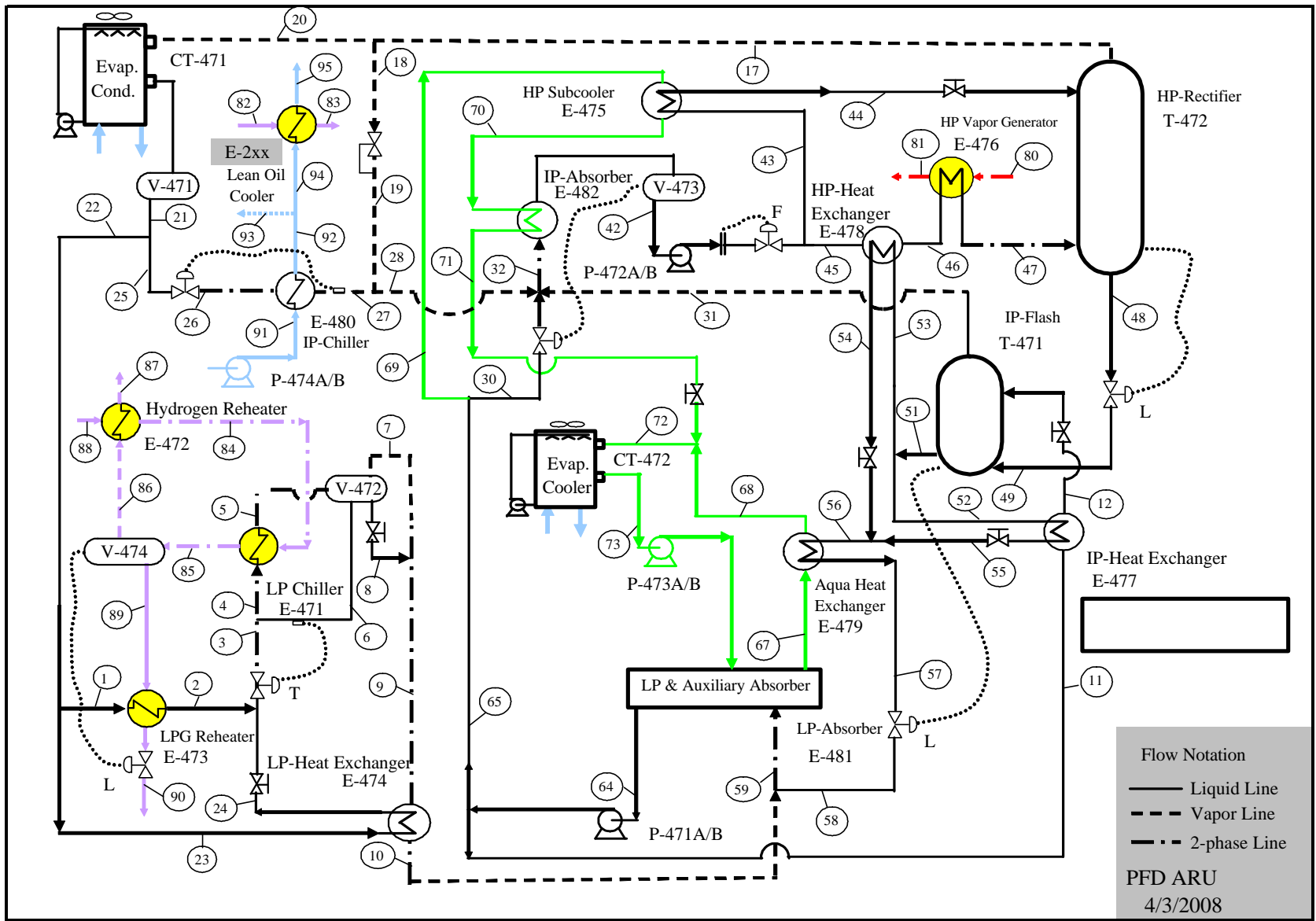


Figure 1 – PFD of Western Refining Co. ARU

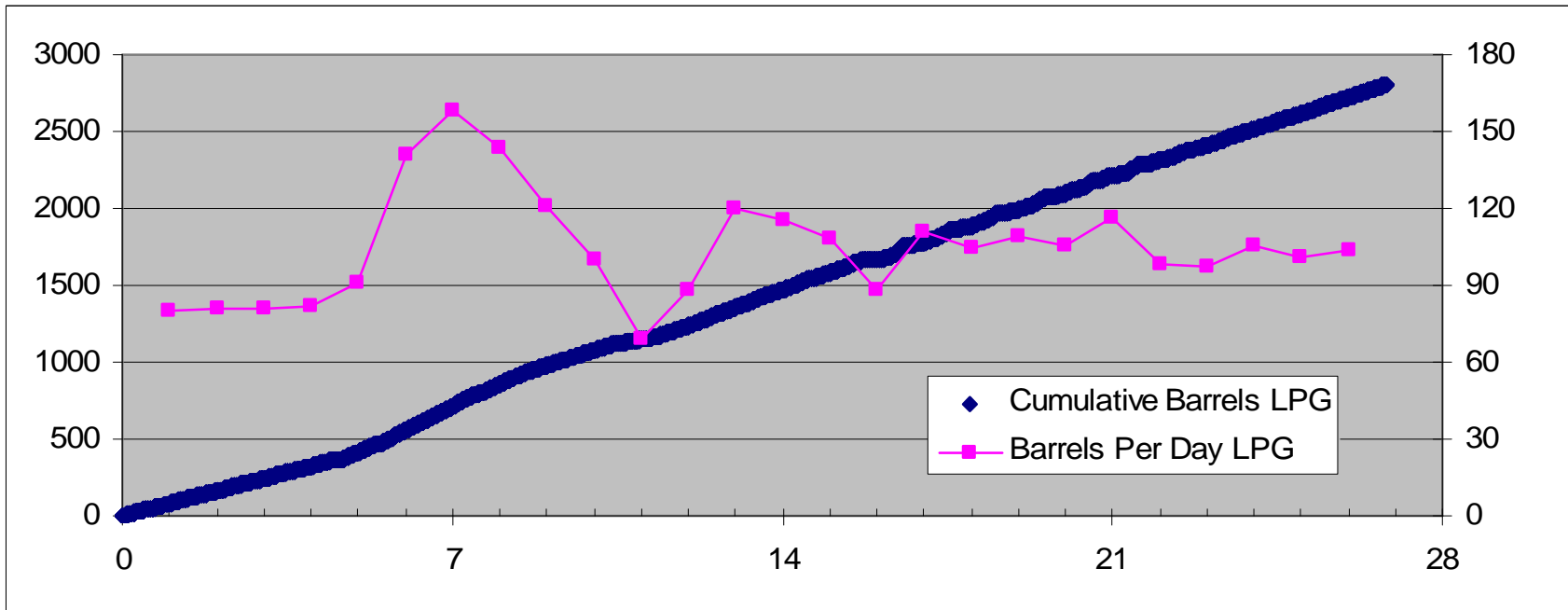


Figure 2 – Western Refining Co. ARU LPG Recovery November 2007

Table 1 – Western Refining Co. ARU Procidia Data 110807b – 8 PM

Date	Time	FT470	FT-471	FT474	FT475	FT476	FT477	LT475	PT474	PT475	PT486	PT507	TI471	TI475	TI476	TI482	TI483	TI484	TI489	TI508
8-Nov-07	7:00:01 PM	15.29	18.2	17.77	2.45	4.83	3.52	48.17	141.84	0.05	28.43	139.53	354.89	287.08	159.86	87.32	62.03	109.27	176.36	-21.78
8-Nov-07	7:05:01 PM	14.74	18.16	16.91	2.46	4.66	3.53	44.48	141.93	-0.1	27.3	139.56	354.57	287.64	160.31	87.15	61.92	109.42	175.62	-22.1
8-Nov-07	7:10:01 PM	15.57	18.03	16.67	2.45	4.69	3.53	41.18	141.53	-0.17	26.33	139.22	353.98	286.62	159.71	86.92	61.87	109.46	175.41	-22.52
8-Nov-07	7:15:01 PM	15.49	18.03	16.97	2.46	4.75	3.54	36.37	141.47	-0.33	27.05	139.19	353.53	285.8	159.69	86.74	61.78	109.25	179.69	-22.87
8-Nov-07	7:20:01 PM	15.59	17.99	17.22	2.47	4.28	3.54	31.4	142.14	-0.15	27.37	139.88	353.46	283.87	161.99	86.9	61.66	109.03	177.44	-22.83
8-Nov-07	7:25:01 PM	16.57	17.34	17.33	2.46	2.82	3.5	39.34	142.01	0.03	26.76	139.75	354.44	283.33	162.22	87.26	61.53	109.4	177.69	-22.35
8-Nov-07	7:30:01 PM	16.54	17.96	16.37	2.46	4.32	3.51	48.94	141.94	0.06	27.29	139.65	355.31	283.46	161.98	87.45	61.38	109.82	175.8	-22.15
8-Nov-07	7:35:01 PM	15.76	17.98	16.48	2.47	4.86	3.54	44.67	142.54	0.01	27.23	140.28	355.78	284.56	159.46	87.37	61.18	109.59	175.49	-22.31
8-Nov-07	7:40:01 PM	15.67	18.08	16.55	2.46	4.91	3.54	38.25	141.99	0	27.69	139.73	355.38	284.62	157.11	87.34	61.02	109.76	174.98	-22.47
8-Nov-07	7:45:01 PM	15.35	17.97	16.02	2.48	4.78	3.55	32.37	143.35	0.1	27.8	141.12	354.61	282.93	160.97	87.44	60.87	109.62	175.06	-22.27
8-Nov-07	7:50:01 PM	15.5	18	17.61	2.47	3.12	3.54	33.83	143.69	0.32	27.95	141.47	354	281.97	160.65	88.02	60.63	109.96	174.37	-21.82
8-Nov-07	7:55:01 PM	15.41	18.11	18.12	2.46	3.3	3.51	44.62	142.8	0.46	28.48	140.58	353.98	281.65	160.43	88.16	60.44	109.86	174.88	-21.43
8-Nov-07	8:00:01 PM	15.64	18.15	16.88	2.42	4.59	3.49	48.21	139.87	0.28	27.51	137.62	353.66	282.35	156.37	87.68	60.3	110	176.3	-21.57
8-Nov-07	8:05:01 PM	15.23	17.95	16.52	2.44	4.67	3.51	44.37	140.35	-0.07	26.88	138.09	354.28	282.68	153.6	86.48	60.07	109.25	171.86	-22.42
8-Nov-07	8:10:01 PM	15.84	17.86	17.15	2.47	4.61	3.53	41.04	142.19	-0.03	26.59	139.87	355.24	283.95	155.53	86.46	59.76	108.95	174.48	-22.68
8-Nov-07	8:15:01 PM	16.47	17.91	16.2	2.47	4.63	3.54	37.67	143.09	0.01	26.84	140.78	355.12	285.19	155.32	86.86	59.54	109.09	172.33	-22.66
8-Nov-07	8:20:01 PM	16.31	18.01	16.78	2.47	4.67	3.54	32.84	142.89	0.02	27.11	140.58	354.37	285.18	155.83	86.88	59.32	109.19	175.71	-22.79
8-Nov-07	8:25:01 PM	15.33	18.14	17.6	2.46	2.69	3.51	35.42	142.19	-0.08	27.05	139.86	353.31	284.98	154.97	86.71	59.03	109.06	173.64	-23.12
8-Nov-07	8:30:01 PM	15.75	18.03	16.97	2.46	4.18	3.51	47.99	142.2	-0.06	27.07	139.89	353.44	284.67	155.78	86.58	58.75	108.77	174.47	-23.13
8-Nov-07	8:35:01 PM	16.52	18.02	16.98	2.45	4.69	3.52	44.43	141.43	-0.13	26.74	139.08	354.15	285.25	154.57	86.55	58.49	108.72	177.52	-23.19
8-Nov-07	8:40:01 PM	16.51	17.94	17.08	2.46	4.7	3.52	41.02	141.45	-0.13	27.22	139.12	354.84	285.28	154.83	86.61	58.03	108.92	173.98	-23.18
8-Nov-07	8:45:01 PM	15.58	17.93	16.43	2.47	4.77	3.54	37.79	141.95	-0.04	26.5	139.7	354.69	284.62	155.68	86.95	57.53	108.95	172.88	-23.08
8-Nov-07	8:50:01 PM	15.67	18.02	16.89	2.48	4.82	3.55	33.82	143.01	0.06	28.11	140.76	354.06	282.4	160.48	87.34	57.11	109.2	173.39	-22.92
8-Nov-07	8:55:01 PM	15.22	18.14	18.56	2.46	3.31	3.53	32.46	142.46	0.14	29.24	140.26	353.63	280.87	159.67	87.44	56.8	109.31	175.05	-22.76
8-Nov-07	9:00:01 PM	15.4	18.08	17.29	2.47	3.7	3.51	46.56	142.5	0.31	28.72	140.3	353.66	279.6	158.3	87.35	56.61	109.11	175.95	-22.36
	ave745:845	15.77	17.95	16.88	2.46	4.28	3.53	40.08	142.11	0.053	27.34	139.85	354.52	283.57	158.87	87.24	60.78	109.51	175.41	-22.29
	ave700:900	15.72	18.00	17.01	2.46	4.29	3.53	40.29	142.11	0.022	27.41	139.83	354.34	284.02	158.21	87.12	59.91	109.32	175.21	-22.51

FT – Flow Transmitter; LT – Level Transmitter; PT – Pressure Transmitter; TI – Temperature Indicator

Table 2 - Western Refining Co. ARU Data Record

	Inst. #	EES St. pt.	110807a 530PM		110807b 8 PM		111507 2 PM	
			Manual	Procidia	Manual	Procidia	Manual	Procidia
HP Rect	P474	17	142.7	142.19	142.4	142.1	130.4	131.15
HP Rect Vap	T476	17	170	169.46	161.7	158.87	183.3	182.78
Unifiner	T471	80	353.9	353.89	355.6	354.52	340	342.97
HP Letd	T475	48	287	288.63	284.4	283.57	274.4	275.38
IP Rect Vap	T489	31	168+-	179.95	188	175.41	186	186.08
HP Pump	T484	42	113.6	112.92	101.1	109.51	106	105.78
HP Pump	F471	42	18.3	17.9	18.2	17.95	18	18
LP Pump	T482	64	90.8	90.54	87.4	87.24	85.4	85.24
Spray Recv	P475	64	0.95	0.969	0.08	0.053	-1.13	-1.08
IP Letd	F474	57	20+-	16.98	19+-	16.88	16.12	15.82
IP Recv	P486	42	32+-	28.68	30.6+-	27.34	12.03	11.84
IP Absorbent	F470	30	10+-	16.1	10+-	15.77	11.6	14.8
Cond Recv	P507	21	139.7	139.78	140	139.85	130	129.96
LPE Ref in	T483	3	64.1	63.91	61.2	60.78	60.5	60.69
LPE ref	F477	22	3.74	3.74	3.54	3.53	3.13	3.12
HC-propane	F476	90	4.3	4.15	4.87	4.28	1.8	1.8
IPE ref	F475	25	2.43	2.46	2.46	2.46	0	0
HC Sep	L475		40	41.2	40	40	40	40
LPE HC	T508	85	-17.8	-18.02	-22.4	-22.29	-24.3	-23.94
Glycol Recv	L478						2B*	
Glycol	F481	91	12.8		12.26		14.6	
IPE Gly out	T503	92	36(32)		35(25)		42	
IPE Ref in	T501	26	22(22)		20(18)		-2	
IPE Gly in	T502	91	36(41)		30			
Gly out	P491	91	64		62		67	
Gly Recv	P493		47		45		50	
IP Recv	L472		7.5B		9B		8B	
Cond Recv	L477		36B		28B		2B top	
IP Rect letd out	T498	51	188(186)		174+-		188	
HP Column	T499	44	90(90)		86(87)		85	
IP Column	T497	12	178(178)		180(179)		170	
IP Column	F472	11	1.11		1.0			
IP Letd out HP HX	T493	54	142(105)		138(138)		125	
HP Pump	T491	45	114		109(109)		106	
Aqua Cooler out	P		54		50		48	
IP Rect	L474		0		0		1.5B	
RHX Liq out	T514	24	19		19		20	
IP HX letd out	T496	55	144		>150		120	
LP Sep out	T515	9	-2		-3		-18	

* Bolts

Table 2 - Western Refining Co. ARU Data Record (cont.)

	Inst. #	EES St. pt.	110807a 530PM		110807b 8 PM		111507 2 PM	
			Manual	Procidia	Manual	Procidia	Manual	Procidia
LP Abs	P496	61						
Aqua HX out	T494	68	112(103)		104(94)		100	
RHX Vap out	T- RHXO	10	(1)					
Aqua Spray	F480	63	43.9		45.66		49.5	
IPAbs Aqua out	T488	71	113(109)		109(109)			
IP Abs in	T486	32	109(122)		85			
LPAbs aqua	F478	66	34		48.2			
IP Letd to LPAbs	T495	57	100(102)		92(93)		89	
Cond Recv	T517	21	74(80)		74(79)		74	
HP Pump out	P483/4		206		205		72	
HP filter out	P487	45	206		201		62	
LP Pump out	P481/2		75		73			
LP filter out	P480	30	65		60		98	
Aqua Pump out	P510/11		80		65			
Aqua filter out	P508	63	62		60			
LPE HC in	T506	84	35?				28	
Liquid HC out	T510	90	54(59)		51(53)		46	
Unifiner to HPG	T470	80	345		345		330	
Unifiner HPG O	T472	81	252		250		255	
HP Gen in	T473	46	185		175		170	
LPE Ref in	T511	4	5		6		-30	
HC Sep	L475		9"		8"		1ft	
Unifiner HPG in	P470	80	290?		295		280	
Reformer gas in	T504	88	96		95		97	
Reformer gas out	T505	87	66		63		64	
HP Gen out	T474	47	298		294		285	
HC to LPG Sep	T507	85	-18		-23		-24	
LPG Sep out	T509	89	-8		-12		-16	
HC Sep	P494	85	295		295		280	
HP Rect	P472	17	144		145		125	
LPA Abs aqua in	T477	66	70		70		70	
LPAbs aqua out	T478	67	89		84		85	
LP Abs out	T480	62	98		92		95	
IP Recv	P485	42	31		30		110	
IP Abs aqua in	T487	70	95		93		90	
Spray Recv	L471		12B		12B		½	
IPAbs out	T485	42	112		110		108	
IPE Ref out	T500	27	58?		30-40		100	
Cond Recv	P506	21					132	
IP Rect	P490	31			31		18	

Table 3 - Western Refining Co. ARU Analysis Summary

Run		110807a	110807b	111507
Temp [F]	Reformer gas in	96.0	95.0	97.0
	Reformer gas out	66.0	63.0	64.0
	LPG condensate	-18.0	-22.3	-23.9
	LPG out	54.0	53.0	46.0
	EG in	38.8	49.4	~
	EG out	17.3	15.6	~
	Unifiner in	353.9	354.4	343.0
	Unifiner out	252.0	250.0	255.0
	Condenser NH3 out	74.0	74.0	74.0
	Cooler Aqua in	105.4	98.8	95.3
	Cooler Aqua out	70.0	70.0	70.0
Flow Rate	HP pump [gpm]	17.9	18.0	18.0
	LP_IPA pump [gpm]	14.8	14.9	17.4
	IP Feed [gpm]	5.3	5.0	1.1
	LPABS Aqua [gpm]	67.3	96.5	83.3
	IP Letdown [gpm]	17.0	16.9	15.8
	LP NH3 [gpm]	3.7	3.5	3.1
	IP NH3 [gpm]	2.5	2.5	0.0
	LPG [gpm]	4.2	4.3	1.8
	EG [gpm]	12.8	12.3	0.0
LP HHX-2p-out Quality		0.970	0.980	0.940
IP Letd-out Quality		0.100	0.090	0.080
Pressure [psig]	HP	139.8	139.8	130.0
	IP	28.7	27.3	11.8
	LP	1.0	0.1	-1.1
Concentration	Refrigerant	0.981	0.987	0.968
	HP-Pump	0.349	0.350	0.260
	LP-Pump	0.258	0.258	0.250
	IP Letdown	0.153	0.158	0.157
	HP Letdown	0.141	0.152	0.164
Performance	LP RT	52	50	39
	IP RT	12	18	0
	COP	0.344	0.372	0.278
Heat Duty [kW]	Evap Condenser CT-471	323.4	305.1	170.0
	Evap Cooler CT-472	541.3	563.4	462.0
	LP-HX E-474	11.7	10.2	13.7
	LPG Reheater E-473	14.5	15.7	6.1
	LP Chiller E-471	182.2	175.3	139.0
	LP ABS E-481	152.4	143.5	105.0
	Auxiliary Abs V-475	123.7	107.9	103.0
	HP-HX E-478	109.1	101.4	112.0
	LP-HX E-477	118.1	120.5	33.7
	Aqua HX E-479	153.2	174.9	239.0
	HP Subcooler E-475	19.3	18.7	14.7
	IP Chiller E-480	42.8	64.5	0.0
	IP Abs E-482	93.1	118.9	0.0
	HP Gen E-476	653.9	644.1	499.0