



Journal Homepage: -www.journalijar.com
**INTERNATIONAL JOURNAL OF
 ADVANCED RESEARCH (IJAR)**

Article DOI:10.21474/IJAR01/ 9250
 DOI URL: <http://dx.doi.org/10.21474/IJAR01/9250>



RESEARCH ARTICLE

HEAT INTEGRATION OF HYDROGEN PRODUCTION FROM GLYCEROL REFORMING.

Tolulope O. Amlogu¹ And Omolola P. Dayo-Odukoya².

1. Department of Chemical Engineering, University of Lagos Akoka, Lagos State, Nigeria.
2. Department of Chemical and Biological Engineering, University of Sheffield Western Bank, S10 2TG, Sheffield, Great Britain.

Manuscript Info

Manuscript History

Received: 10 April 2019
 Final Accepted: 12 May 2019
 Published: June 2019

Key words:-

Glycerol reforming, Hydrogen production, Bio-diesel, Heat integration, Pinch analysis.

Abstract

Glycerol is a major by-product of bio-diesel, making up at least 10% of the biodiesel produced. In a world concerned about sustainability and recycling, this project aimed to find an alternative, cost-effective source of energy (hydrogen) from glycerol, a waste product of bio-diesel, via the reforming process. This study examined the use of the 'Pinch technology' as a technique in conserving energy in the glycerol reforming process for hydrogen production. The pinch analysis was carried out on two methods of reforming namely: Aqueous phase reforming (APR) and Steam reforming (SR). The results obtained showed that the unit production cost of H₂ (in \$/kmol), before and after integration for APR are 31.68 and 31.56 consecutively and for SR- 48.7 and 52.99 consecutively. Energy recovered from APR and SR are 29, 820kW (86%) and 37,400kW (69%) respectively. These energy savings made, reduced the operating cost by 92% for APR and 75% for SR but an increased capital cost was incurred as a result of the additional heat exchangers that were required to achieve recovery. In this study, APR was found to be a more cost-effective and energy saving method (in both the non-integrated and integrated case), which makes it very ideal in our energy-dependent world, because it is carried out in liquid phase and at low or slightly elevated temperature.

Copy Right, IJAR, 2019,. All rights reserved.

Introduction:-

The demand for clean, renewable and sustainable energy has been on the rapid increase due to the depleting crude oil reserves and the negative impact unclean energy sources such as fossil-based fuel have on the atmosphere, the water cycle, and the general well-being of humans on planet earth¹. These issues have led to various researches centered on discovering economically viable and environmentally friendly (clean) energy sources and as a result, new technologies requiring the use of renewable feed stock has been the focus of intense process developments within the past few decades². The discovery of bio-fuel (fuel derived from biomass) was a major breakthrough in the search for clean energy. Currently, a few countries like the US are already making use of renewable feed stocks derived from biomass for production of diesel. According to the U.S Energy Information Administration (EIA) in 2016, diesel was a major part of the about 4.8% of biomass energy consumed in the U.S. Unfortunately this bio-fuel cannot compete favorably with fossil based fuel as a result of the economy of its production, that is, production cost³. According to the US Department of Energy, Alternative Fuel Data Centre, 2014, the cost of biodiesel is about

Corresponding Author:-Tolulope O. Amlogu.

Address:-Department of Chemical Engineering, University of Lagos Akoka, Lagos State, Nigeria.

12% higher than the price of diesel. If biodiesel must compete with diesel effectively, alternative routes for reduction in production cost must be sought for.

Meanwhile, advancement in the production of biodiesel has led to the tremendous increase in supply of glycerol, being the major by-product of biodiesel. About 10% per unit mass of biodiesel produced is glycerol⁴. Although glycerol has been useful in other industries such as; food, beverages, body care, pharmaceutical etc., its rate of production far outweighs the demand for it in these industries. This has led to the oversupply and market flooding of glycerol, hence leading to drop in demand and market value (price) of glycerol. Over the years, it has been clearly observed that the amount of glycerol produced exceeds the actual consumption, and the mismatch is increasing further². The U.S Department of Energy (DoE) estimates that if the United States were to produce enough biodiesel to displace only 2% of the current petroleum diesel usage, an additional 800 million pounds of glycerol would be produced per year.

Hydrogen (H₂) is a very attractive form of energy found useful in the chemical processing, transportation, petroleum, food and beverage industries amongst others⁵. Over the years, hydrogen has been produced by reforming hydrocarbon- natural gas (fossil fuel). This implies that a non-renewable fuel is being burnt in order to produce hydrogen. With so many concerns about 'human activities putting the planet in danger', this is not a recommended route for hydrogen production and can be greatly avoided by reforming glycerol, an oxygenated hydrocarbon which is from a renewable source⁶.

The Aqueous Phase Reforming (APR) is the only method in which the reforming is carried out in liquid phase at low or slightly elevated temperature. H₂ can be generated without previous feed (glycerol and water) vaporization, therefore it requires less temperature and energy is conserved⁷. Steam reforming (SR) on the other hand is the oldest and most common method of reforming in industries⁸. It involves the splitting of hydrocarbons in the presence of steam at high temperature. It is described by the equation:



This method has high requirement for energy because it operates at very high temperature between 700 to 900°C⁹.

Heat integration also known as pinch technology is the systematic and general method for analysis, design and optimization of energy in a production system. It emphasizes on the efficient use of energy. Needs of opposite kinds are combined to enable savings. The field was birthed by the concept of heat recovery pinch guided by the principle of heat transfer with temperature difference as driving force¹⁰. It involves the merging of cold streams with hot streams, condensation with evaporation, identifying near optimal level of heat recovery and designing heat exchanger. Heat could be recovered and reused from one unit operation to another either by a direct process or an indirect process.

In view of all these, this research study was carried out using Aspen PLUS simulator to develop a cost effective method for hydrogen production from glycerol thereby improving the overall economy of the biodiesel production by;

1. developing and comparing 2 different methods of glycerol reforming that will be for cost-effective production of hydrogen and;
2. heat integration of the reforming processes developed, to minimize hydrogen production cost over a period of time.

Experimental Section

The Aspen PLUS simulator was used for the process development of the two reforming methods adopted: the Aqueous Phase Reforming (APR) and Steam Reforming (SR).

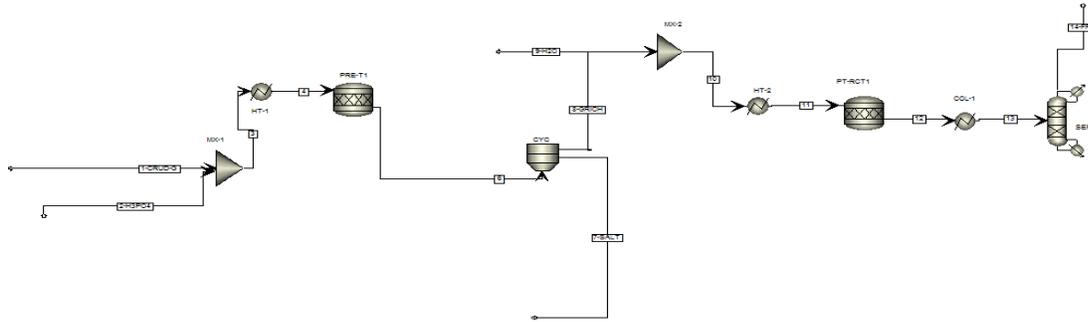
Process Simulations

Pretreatment stage.

The first process design was the pretreatment of the crude glycerol with acid⁷. Scheme 1 shows the flow sheet for the pretreatment process. Phosphoric acid (H₃PO₄) - 25% v/v was used. H₃PO₄ reacts with KOH in the crude glycerol to produce solid potassium salt as shown in the reaction in equation 2:



The solid potassium salt is separated by decantation. The stream flows into another reactor where hydrolysis of triolein takes place to produce fatty acid (oleic acid) and glycerol. The stream moves to the separator where oleic acid is separated from pure glycerol.

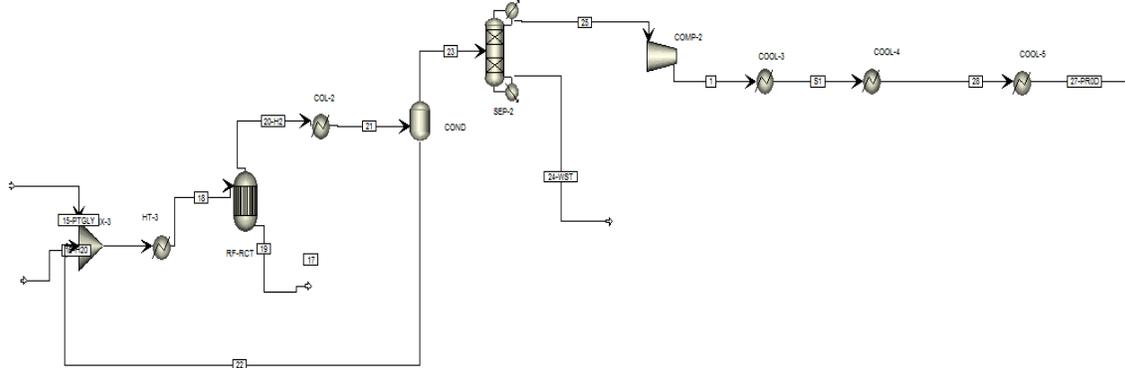


Scheme 1:-Process flowsheet for pretreatment.

Aqueous phase reforming (APR) process simulation

The property model Peng-Robinson was selected. The equilibrium reactor was assumed to model a heterogeneous system. Two reactions takes place in the reformer, one is the decomposition of glycerol and the other is the water gas shift reaction (WGS) which takes place in the ultra shift zone.

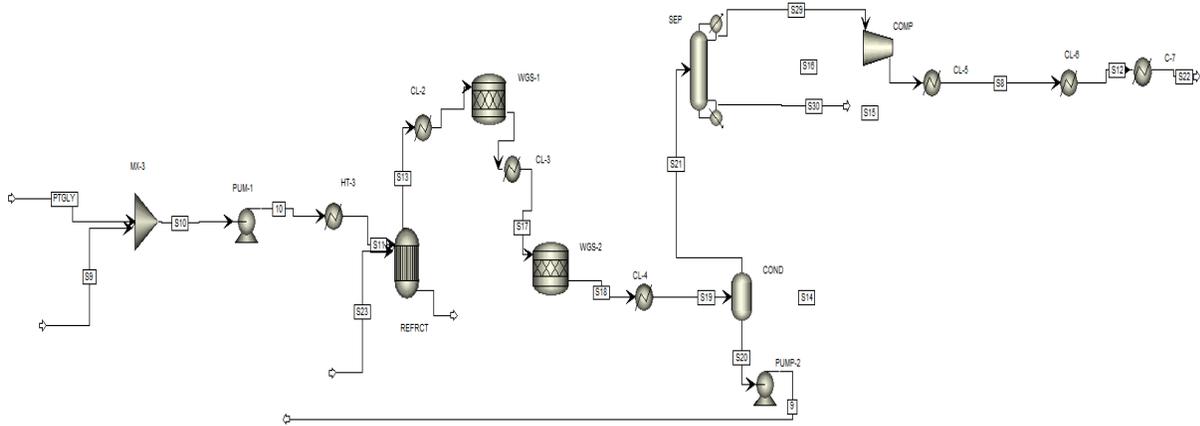
The unit operations, plant capacity and input conditions for this case that is, reforming of glycerol, were selected based on the research carried out by Davada et al⁵. The reactor temperature was set at 498K (225°C) with a pressure of 38bar. The process flow sheet is shown in scheme 2.



Scheme 2:-Process flow sheet for aqueous phase reforming.

Steam Reforming (SR) Process Simulation

The components for SR were selected, with property method GLYCOL for simulation. Process conditions for simulation were selected based on research carried out by Molburg et al¹². The reactor temperature was set to 745°C. The flow sheet is shown in scheme 3 below.



Scheme 3:-Process flow sheet for steam reforming.

Heat Integration (Pinch Analysis)

Heat integration was carried out with the following steps;

Obtained relevant data: Data such as heat load and temperature for all process streams and utilities were extracted from flow sheet of the simulated process design. The hot and cold streams were also identified. The necessary data for each process stream are the following: T_s (source temperature, °C); T_t (target temperature, °C) and; CP (heat capacity flow rate KW/°C).

Generated energy and utility targets:

1. DT_{min} is the minimum temperature difference between the hot composite curve and the cold composite curve at the pinch point (the point of closest approach between the two curves). It represents a tradeoff between the capital investment (which increases for smaller values of DT_{min}) and the operating cost (which reduces for smaller DT_{min}). The optimum DT_{min} was obtained by plotting total cost targets against DT_{min} .
2. Determine targets Q_{cmin} (minimum external cooling utility required) and Q_{hmin} (minimum external heating utilities required) by plotting composite curves. The heat load or enthalpy (ΔH) was computed using the formula below and then plotted against temperature.

$$\Delta H = CP (T - T_{ref}) \quad (T_{ref} - \text{the reference temperature})$$

T represents T_s or T_t depending on what parameter is being calculated

Designed heat exchanger network: The heat exchanger network is designed and represented in a grid diagram by matching streams. The system is divided into 2 subsystem; 1 above pinch (higher temperature range) and the other below pinch (lower temperature range). The design starts at the pinch point and gradually moves away from the pinch making sure hot streams are utilized above pinch and cold stream utilized below pinch. The matching rules for pinch exchangers (those situated immediately above or below pinch) can be expressed mathematically by;

Above the pinch point, $CP_h \leq CP_c$; below the pinch point $CP_h \geq CP_c$

Where CP_h represent heat capacity flow rate for hot stream and

CP_c represent heat capacity flow rate for cold stream

If these rules are not satisfied then stream splitting would be required^{10, 13}.

Costing

Cost estimation of both reforming methods was carried out and its cost effectiveness determined. Aspen Economic analyser in combination with other method was use for cost estimation of the reforming processes.

Factorial Method

The fixed capital cost of the project is given as a function of the total purchase equipment cost by the formula¹⁴.

$$C_f = f_L \times C_e$$

Where: C_f =fixed capital cost; C_e =the total delivered cost of all the major equipment items; f_L =the “Lang factor” (dependent on the process) = 3.63 for fluids processing plant.

The cost factors that make up the Lang factors are broken down individually as seen in Table 1

Table 1:-Typical factors for fixed capital cost estimation (extract from ¹⁴)

Items	Process Type
1 Major equipment, total purchase cost	PCE
F ₁ - Equipment erection	0.45
F ₂ -Piping	0.45
F ₃ -Instrumentation	0.15
F ₄ -Electrical	0.10
F ₅ -Buildings, process	0.10
F ₆ -Utilities	0.45
F ₇ -Storage	0.20
F ₈ -Site development	0.05
F ₉ -Ancillary Buildings	0.20
PPE ($\sum F$)	3.15
F ₁₀ Design and Engineering	0.25
F ₁₁ Contractor's fee	0.05
F ₁₂ Contingency	0.10
($\sum F_{10-12}$)	1.40

Total Physical Plant cost (PPC) = Installed Cost x $\sum F$

Fixed Capital = PPC x F₁₀₋₁₂

Heat Exchanger Cost

The cost of heat exchangers:

Capital cost = a + b (A/No. of shells)^c x No. Of shells

Where A= Heat transfer area; a, b and c are constants with values 1000, 800 and 0.8 respectively.

Variable Cost and Fixed Cost

The variable and fixed operating cost can calculated using information from Table 2 below.

Table 2:-Summary of production cost (extract from ¹⁴)

Variable costs	Typical Values
1 Raw materials	From flow- sheets
2 Miscellaneous materials	10% of maintenance
3 Utilities	From flow-sheet
4 Shipping and packaging	Usually negligible
Sub-total A	
Fixed cost	
5 Maintenance	5-10% of fixed capital
6 Operating labour	From manning estimates
7 Laboratory cost	20-23% of operating labour
8 Supervision	20% of (6)
9 Plants overhead	50% of (6)
10 Capital charges	10% of fixed capital
11 Insurance	1% of fixed capital
12 Local taxes	2% of fixed capital
13 Royalties	1% of fixed capital
Sub-total B	
Direct production cost A+B	
13 Sales expense	20-30% of direct production cost
14 General overheads	
15 Research and development	
Sub-total C	
Annual production cost A+ B+ C=	

Results And Discussion:-

3.1 Pretreatment Stage

Table 3 shows the composition of glycerol, before and after pretreatment.

Table 3:-Composition of glycerol, before and after pretreatment

Content	Before Pretreatment (%) (composition)	After Pretreatment (%) (composition)
Glycerol	60	84
Water	10	11.81
Triolein (Unreacted Triglyceride)	15	0
KOH	5	0.23
Methanol	10	3.96

3.2 Aqueous Phase Reforming

Cost evaluation for non-integrated APR case

In table 4, the cost of raw material required for the production of 2,000,000kmol H₂ per annum which amounts to \$22,272,920 has been shown, as well as the utility cost and rates (approximately \$812 per hour) which is represented by table 5 and the summary of total production cost of APR shown in table 6. These calculations show that it will cost approximately \$32 per kmol of H₂.

Table 4:-Cost of raw materials required for production of 2million kmol H₂ per annum

Name	Specification	Price (\$)	Rate/yr	Cost(\$/yr)
Glycerol	Crude	0.11023/kg	8.5607 x10 ⁷ kg	9.43652x10 ⁶
Phosphoric acid	99%	25.448/L	1.9 x 10 ⁶ L	1.20371x10 ⁷
Mains Water (process water)	-	0.00155/L*	9.62 x10 ⁹ L	4.56 x10 ⁴
Catalyst(Pt/Al ₂ O ₃)	5 wt %	10,324/kg	73kg	7.537 x 10 ⁵
Total				2.227292 x 10⁷

*Cost of water is given as seen in the industrial water pricing for OCED countries.

Table 5:-Utility rates and cost

Name	Fluid	Rate (Btu/hr)	Cost (\$/hr)
Electricity		746.374 kW	57.843985
AP-UTIL-U-3	Steam	505,717.5	-1.330037
AP-UTIL-U-4	Steam	904,565.2	-1.800085
AP-UTIL-U-UCLN33	Propane	99,788,940	288.390037
AP-UTIL-U-CLN5	Propane	798,417.1	2.307425
AP-UTIL-U-HT1	Steam	506,624.3	1.013249
AP-UTIL-U-HT2	Steam	6,580,769	24.283038
AP-UTIL-U-HTN1	Steam	114,402,900	422.146701
AP-UTIL-U-VL	Propane	6,463,417	18.679275
Total			811.533588

The remaining generation of high energy is considered as service in ASPEN hence the negative sign.

Table 6:-Total Production Cost for APR

		Cost (\$)
Fixed Capital		29,800,645.00
Fixed Operating Cost	Taxes	447,009.68
	Insurance	447,009.68
	Maintenance	2,980,064.50
Variable Operating Cost	Utilities	7,109,034.21
	Raw Materials	22,272,920.00
	Miscellaneous Operating	298,006.45

	Material	
Total		63,354,689.52

Therefore the total manufacturing cost of 2million kmol H₂ is \$63,354,689.52 per annum

$$\begin{aligned} \text{Production cost of H}_2/\text{kmol} &= \frac{63,354,689.53}{2,000,000} \\ &= \$31.68\text{kmol}^{-1} \end{aligned}$$

Pinch Analysis for APR

Stream data required for pinch analysis such as inlet and outlet temperature, enthalpy were extracted from the flow sheet. The data is shown in the table 7.

Table 7:-Stream Data for APR

Name	Type	T _s (°C)	T _t (°C)	CP (kJ/C.h)	Enthalpy(kW)	Segment
1	Hot	329.8372	25	153,862.6771	2,199	Hot
2	Hot	270	25	-	1,894	Hot
3	Cold	57	270	-	1,929	Cold
4	Cold	47.2	255	-	3.353 x10 ⁴	Cold
5	Hot	255	100	-	2.925x10 ⁴	None

Determining Energy and Utility Targets

The composite curve is plotted with stream data in order to determine the energy targets. From the plot in figure 1, it is observed that the optimum DT_{min} (minimum temperature that will give us the maximum energy recovery) is 6°C.



Figure 1:-Optimum DT_{min} plot for APR

The optimum DT_{min} is used in plotting the composite curve as seen in figure 2. The amount of energy recoverable from the process as shown by the region of overlap on the composite curve is 29,820kW. The energy targets QC_{min} and QH_{min} are seen in the plot as 3,519kW and 5,787kW respectively. The pinch temperature was obtained from the curves as 225.4°C. The summary is given in table 8.

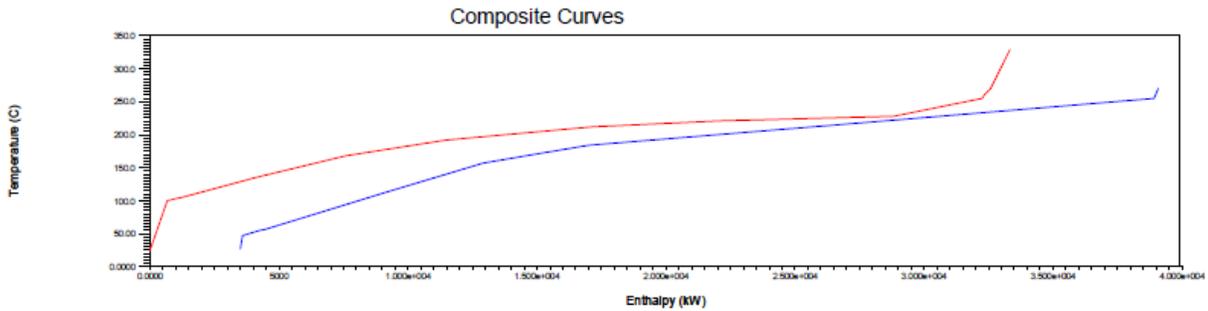


Figure 2:-Composite curve for APR (*Indicators: Red – hot; Blue - cold)

Table 8:-Heat Recovery Summary

	Heating	Cooling
Non-integrated APR Simulation Energy Requirement (kW)	35,607	33,339
Energy Targets (kW)	5,787	3,519
Savings (kW)	2.982×10^4	2.982×10^4
Savings (%)	83.7	89.4

Heat Exchanger Network Design for APR

Capital Cost of heat exchanger network was calculated using data extracted from Aspen PLUS. After pinch analysis was carried out, the grid diagram was designed for integrated APR case. The number of heat exchangers increased to 15 which consist of 3 coolers, 4 heaters and 8 stream to stream heat exchangers. As a result of these, the heat transfer area of the heat exchangers increased to 6081m² with 54 shells as shown in table 9.

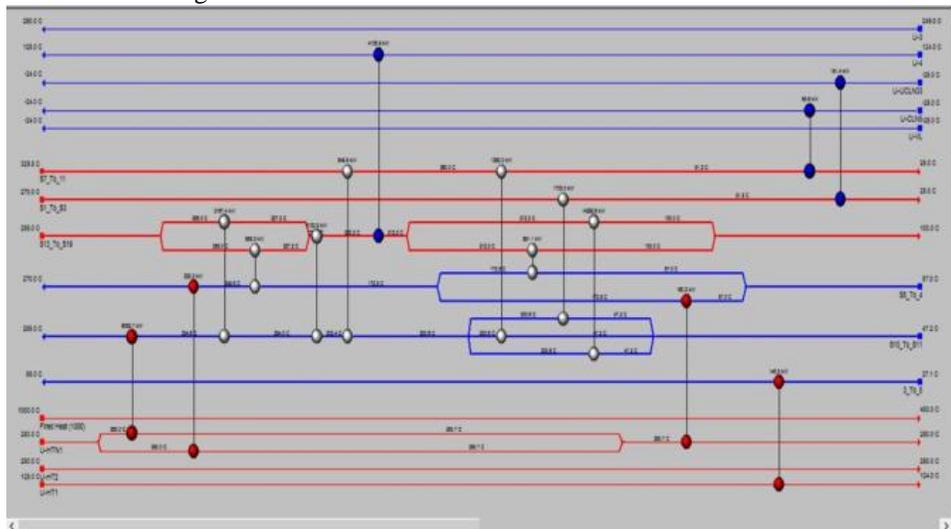


Figure 3:-Heat exchange network for integrated APR case.

The blue circles linked with vertical lines are cooler while the red ones represent heaters, the grey ones represent stream to stream heat exchangers.

Cost Evaluation for APR after Integration

The total utility cost (energy cost), as shown in Table 9, required for plant operation per year is \$7,109,034.21. A saving of 92% was achieved (from operating costs) through heat integration which amounts to \$6,563,297.14.

Table 9:-Summary of Energy Cost

	Non-integrated Case	APR	Integrated APR Case	Deficit/Surplus
No. of units	8		15	
No of Shells	15		54	
Area (m ²)	3,432		6,081	
Heat Exchanger Capital Cost (\$)	6,471,000		12,798,854.91	6,327,854.91 (198%)
Operating Cost (\$/yr)	7,109,034.21		545,737.09	6,563,297.12
Operating Cost Savings (%)	0		92	
Total Production Cost (\$)	63,354,689.52		63,119,247.31	
Unit Production Cost of H ₂ (\$/kmol ⁻¹)	31.69		31.56	

The capital cost for the integrated APR case is higher, due to the increased heat transfer area, but a total of 92% was saved from operating cost (utilities). Therefore the unit cost of production of H₂ after heat integration is:

$$\begin{aligned} \text{Production cost of H}_2/\text{kmol} &= \frac{63,119,247.31}{2,000,000} \\ &= \$31.56\text{kmol}^{-1} \end{aligned}$$

Steam Reforming

The cost evaluation for the non-integrated SR case was carried out just as it was done for the APR case above using data extracted from Aspen Economic Analyzer. Table 10 shows the breakdown of production cost for SR.

Table 10:-Total Production Cost for SR

		Cost (\$)
Fixed Capital		34,100,057.11
Fixed Operating Cost	Taxes	511,500.86
	Insurance	511,500.86
	Maintenance	3,410,005.71
Variable Operating Cost	Utilities	36,306,520.80
	Raw Materials	22,272,920.00
	Miscellaneous Operating Material	341,000.57
Total		97,453,505.91

The total manufacturing cost of 2million kmol H₂ is \$97,453,505.91 per annum. Hence:

$$\begin{aligned} \text{Production cost of H}_2/\text{kmol} &= \frac{97,453,505.91}{2,000,000} \\ &= \$48.73\text{kmol}^{-1} \end{aligned}$$

Pinch Analysis for SR

Stream data shown in Table 11, such as inlet and outlet temperature (°C) and enthalpy (kW) were extracted from the flow sheet, the heat capacity flow rate (kJ/C.h) was calculated using Aspen Energy Analyzer.

Table 11:-Process Stream Data for SR

Stream	Type	Ts(°C)	Tt(°C)	CP (kJ/C.h)	Enthalpy (kW)	Segment
1	HOT	745.437	350	118,683.1	13,036.58	None
2	HOT	270	25	-	1,900.241	Hot
3	COLD	47.42801	745.437	-	52,607.23	Cold
4	HOT	204.445	100	-	22,508.51	Hot
5	COLD	56.97533	270	-	1,933.836	Cold
6	HOT	427	204.445	-	14,357.95	Hot
7	HOT	327.6606	25	-	1871	Hot
8	COLD	27.12742	55	19,179.05	148.4916	None

Pinch analysis was carried out in other to obtain the possible energy saving for the SR process.



Figure 4:-Optimum DT_{min} for SR

The composite curve is plotted with stream data in order to determine the energy targets. The optimum DT_{min} is first determined to be 24°C. Figure 5 shows the composite curve.

The region of overlap on the composite curve is 37,400kW, with a pinch temperature of 217.2°C.

The summary of heat recovery is shown in Table 12 below

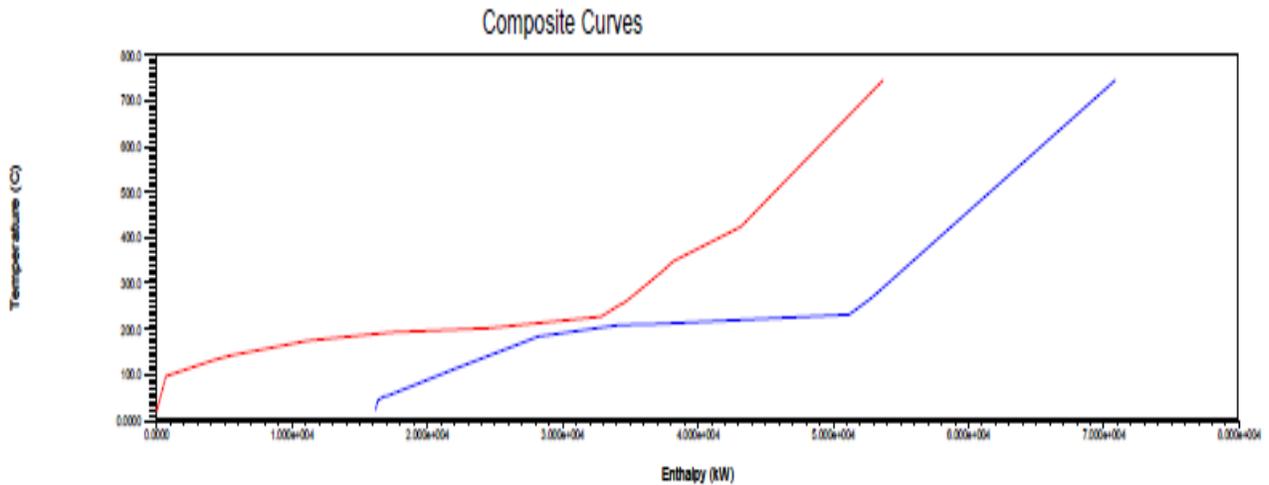


Figure 5:-Composite curve for SR

Table 12:-Heat recovery summary for SR

	Heating	Cooling
Non-Integrated SR Simulation Energy Requirement (kW)	54,680	53,670
Energy Targets (kW)	17,240	16,240
Savings (kW)	37,400	37,400
Savings (%)	68.4	69.7

Heat Exchanger Network Design for SR

After pinch analysis was carried out, the grid diagram was designed for the integrated SR case. The number of heat exchangers increased to 20. As a result of these, the heat transfer area of the heat exchangers increased to 6,903m² with 61 shells.

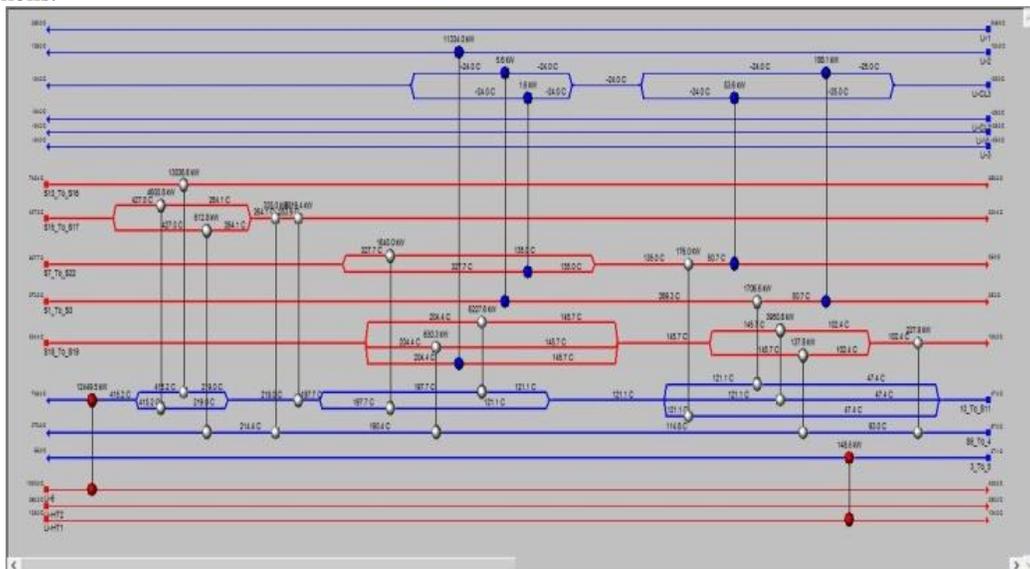


Figure 6:-Heat exchange network integrated SR case

The total utility cost (energy cost), as shown in Table 13, required for plant operation per year is \$36,306,520.80. A saving of 75.2% of operating cost was achieved through heat integration. Capital Cost of heat exchanger network was calculated using data extracted from Aspen PLUS.

Table 13:-Production cost for both SR cases.

	Non-integrated Case	SR	Integrated SR Case	Deficit/Surplus
No. of units	9		20	
No. of Shells	11		61	
Area (m ²)	788		6903	
Heat Exchanger Capital Cost (\$)	7,002,253		42,850,540	35,848,287 (612%)
Operating Cost (\$/yr)	36,306,520.80		9,002,681.31	27,303,839.49
Operating Cost Savings (%)	0		75.2	
Total Production Cost (\$)	97,453,505.91		105,997,953.42	
Unit Production Cost (\$/kmol ¹)	48.7		52.99	

The capital cost for the integrated SR case is higher due to the increased heat transfer area, but a total of 75.2% was saved from operating cost (utilities). Therefore the cost of production of H₂ after heat integration is

$$\text{Production cost of H}_2/\text{kmol} = \frac{105,997,953.42}{2,000,000} \\ = \$52.99/\text{kmol}^{-1}$$

Discussion:-

This comparative study between the heat integration of APR and SR was carried out in order to determine the cost effective method. The plant requires 85.6million kg of glycerol to run per year in order to produce 2million kmol H₂.

Heat integration was carried out in order to maximize energy, by recovering energy from stream to stream matching. The total recoverable amount of energy was determined from the composite curve plot, the hot and cold composite are placed together on a single temperature-enthalpy plot. The hot and cold composite curves represent the sum of all heat sources and all heat sinks respectively within a process. To do this appropriately, the optimum DT_{min} was first determined. The optimum DT_{min} will give us a balance between capital cost and operating cost. It represents a

tradeoff between the capital investment (which increases for smaller values of DT_{\min}) and the energy/operating cost (which reduces for smaller DT_{\min}). DT_{\min} was determined for APR and SR as 6°C and 24°C respectively. Having done this, the composite curve is plotted. The plot showed that a total of 29,820kW (86%) was recovered from the APR non-integrated case, with appropriate stream matching with the aid of the grand composite curve.

Operating cost for the APR of which utilities is a \$7.11million was reduced by 92% to \$0.55million. From the new network design, the integrated case had a total of 15 heat exchangers consisting of 3 coolers, 4 heaters and 8 stream to stream heat exchangers, as a result the new heat transfer area increased to 6081m^2 .

The cost evaluation carried out gave an overall 0.37% savings on total cost of production with $\$31.68\text{kmol}^{-1}$ and $\$31.56\text{kmol}^{-1}$ as the unit cost of H_2 for the non-integrated case and integrated case respectively. The mere 0.37% saving is insignificant and discouraging when compared to the 198% increase in the capital cost of heat exchangers implementation, but a more critical analysis showed that the savings made from the operating cost(utilities) would make up for the increased capital cost over a short period of time.

A similar analysis was carried out on Steam reforming (SR) using the same procedure as in APR above. The composite curve revealed that a total of 37,400kW (69%) was recovered with heat integration. The operating cost was reduced by 75.2% from \$36.31million to about \$9million. The new network design has a total of 20 heat exchangers including coolers, heaters and stream to stream heat exchangers leading to a transfer area of 6903m^2 . An increase in production cost of about 9% with unit production cost of $\$52.99\text{kmol}^{-1}$ of H_2 for integrated case and $\$48.7\text{kmol}^{-1}$ for the non-integrated case SR case.

A massive increase was seen in the capital cost for heat exchangers due to the implementation of the new stream to stream heat exchanger for the maximization of energy in form of utilities.

From results obtained, it is safe to say that heat integration does not necessarily bring about an upfront reduction in production cost but rather production cost is saved over a short period of time through savings from utilities. It is wiser to save operating cost while investing in capital cost because capital cost is fixed and would not need review in the nearest future but operating cost is variable and could change based on different factors such as; government policies, availability etc.

Conclusion:-

The previously wasted glycerol is now found useful for energy production, in the form of hydrogen. Revenue generated from this would improve the biodiesel economy greatly and above all gradually sweep out environmental pollution from fossils fuels as the use of biodiesel increases. This study has proven the fact that aqueous phase reforming (APR) is a more cost-effective and energy saving method (in both the non-integrated and integrated case). With steam reforming there is a high demand for energy and this comes in as a disadvantage to the energy dependent world.

The break even analysis showed that the additional capital cost incurred due to the additional heat exchangers would be gained back from the savings made from the operating cost over the break even period. For the APR, it will take 12 months (1 year) for the surplus (savings) from operating cost to pay for the deficit from heat exchangers capital cost. For the SR the break even period is 16 months.

Acknowledgement:-

I appreciate Dr. Usman Lawal of University of Lagos, Akoka for his support all through and to the completion of this work.

References:-

1. Riddell A. Ronson, S. Counts, G. and Kurt, S. "Towards Sustainable Energy: The Current Fossil Fuel Problem and the Prospects of Geothermal and Nuclear Power," 2015. [Online]. Available: http://web.stanford.edu/class/e297c/trade_environment/energy/hfossil.html. [Accessed 17 July 2016]
2. Avasthi K. S, Reddy R. N, Patel S. (2013). Challenges in the production of hydrogen from glycerol-a biodiesel byproduct via steam reforming process. *Procedia Engineering*, 51: 423 – 429
3. Energy Information Administration, Monthly Energy Review, March 2016, Table 10.3, http://www.eia.gov/totalenergy/data/monthly/pdf/sec10_7.pdf
4. Silvey L. G. (2011). Hydrogen and Syngas Production from Biodiesel Derived Crude Glycerol.
5. Davda R. R, Shabaker J. W, Huber G. W, Cortright R. D, Dumesic J. A. (2005). A review of catalytic issues and process conditions for renewable hydrogen and alkanes by aqueous-phase of oxygenated hydrocarbons over supported metal catalysts. *Applied catalysis B:environmental*, 56: 171- 176.
6. Sabri, F. (2013). Catalysts for hydrogen production by the auto-thermal reforming of glycerol, M.Sc Thesis, University of Regina, Saskatchewan.
7. El Doukkali, M. Iriondo, A. Cambra, J. Gandarias, I. Jalowiecki-Duhamel L. Dumeignil, F. and Arias P.L. (2014). Deactivation study of the Pt and /or Ni-based γ -Al₂O₃ catalysts used in the aqueous phase reforming of glycerol for H₂ production. *Applied Catalysis*, 472: 80-91.
8. Roussi re, T. L. (2013). Catalytic reforming of methane in the Presence of CO₂ and H₂O at high pressure. Ph.D. Dissertation, Karlsruhe Institute of Technology, October 2013.
9. Stelmachowski, M. (2011). Utilization of glycerol, a by-product of the transesterification process of vegetable oils: a review. *Ecological Chemistry and Engineering*, 18 (1).
10. Gundersen, T. (2000). A Process Integration Premier, International Energy Agency. Sintef Energy Research.
11. Nanda, M. R, Yuan, Z. Qin, W. Poirier, M. A and Chunbao, X. (2014). Purification of Crude Glycerol using Acidification: Effects of Acid Types and Product Characterization. *Austin J Chem Eng*, 1(1): 1004.
12. Molburg, J. C, Doctor, R. D. (2003). Hydrogen from steam methane reforming Argonne National Laboratory, 20th Annual International Pittsburgh Coal Conference, 7-19.
13. CANMET Energy Technology Centre- Varennes. (2003). Pinch Analysis: For efficient use of energy water and hydrogen. Natural Resource Canada.
14. Sinnott, R. K. (2005). *Chemical Engineering Design*. Coulson and Richardson's Chemical Engineering Series. Elsevier Butterworth-Heinemann, 6(4): 243-258.